CONFERENCE PROCEEDINGS

INTEGRATING MEA REGENERATION WITH CO_2 COMPRESSION TO REDUCE CO_2 CAPTURE COSTS

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ABSTRACT

Capturing CO_2 from coal-fired power plants is a necessary component of any large-scale effort to reduce anthropogenic CO_2 emissions. Conventional absorption/stripping with monoethanolamine (MEA) or similar solvents are the most likely current process for capturing CO_2 from the flue gas at these facilities. However, one of the largest problems with MEA absorption/stripping is that conventional process configurations have energy requirements that result in large reductions in the net power plant output. Several alternative process configurations for reducing these parasitic energy requirements were investigated in this study with the assistance of the Platte River Power Authority, based on recovering energy from the CO_2 compression train and using this energy in the MEA regeneration step. These configurations included CO_2 vapor recompression heat recovery, and multipressure stripping with and without vapor recompression heat recovery. Four process configurations were simulated using a rigorous rate-based model, and the results were used to prepare capital and operating cost estimates. CO_2 capture economics are presented, and the cost of CO_2 capture (cost per tonne avoided) is compared among the base case and the alternative process configurations. Energy savings of the improved configurations clearly outweighed the modest increases in capital cost to implement them; it is therefore likely that one of these improved configurations would be used whenever MEA-based scrubbing technologies are implemented. These results indicate an improvement to commercial MEA-based technologies which is helping to meet DOE's Sequestration Program targets.

INTRODUCTION

One of the goals of the U.S. Department of Energy's Carbon Sequestration program that is administered within the Office of Fossil Energy and managed by the National Energy Technology Laboratory is to ensure that there are a number of carbon dioxide capture and sequestration technologies available should the requirement for carbon management be needed. Currently there is only one CO_2 post-combustion capture technology that has the luxury of claiming commercial status, Monoethanolamine (MEA) scrubbing systems. Unfortunately they come at a high operational and financial price. The latter does not mean that the Sequestration Program's search for novel technologies excludes advancements of MEA technology based systems. To the contrary, as the only near term option for carbon dioxide capture, incremental improvements in MEA scrubbing systems will be necessary as we transition to other carbon dioxide capture emerging technologies.

IEA stated that "MEA scrubbing, while not inexpensive, is the least expensive of the near term options in terms of cost per unit of CO_2 captured (IEA 1994)." It is also the only CO_2 capture technology that can be and has been used continuously and successfully in post combustion power generation plants. Its track record of operation for either utility or non-utility operation, although not at full load, has also made it the baseline technology to compare all other novel post combustion capture technologies and concepts. This fact supports the aforementioned statement that MEA scrubbing research is necessary.

Nonetheless, there are several significant challenges with using MEA scrubbing on flue gas. Residual oxygen, SO_2, and other species will cause chemical degradation of the MEA. The MEA liquid solution can be corrosive to process equipment. Finally, and perhaps most significantly, the capital and energy costs to implement MEA scrubbing on power plant flue gas are high. The energy costs associated with the amine regeneration step is particularly high and therefore much research has been directed at reducing these costs.

In previous work, researchers at the University of Texas defined the actual energy, ideal/theoretical energy, and lost work involved with MEA absorption and stripping approaches (Rochelle, 2003). The result was that more than half of the energy required by a standard MEA and CO_2 compression approach was the result of lost work; losses in the MEA stripper were the largest. Several innovative processing approaches were proposed to reduce this lost work; in general, these approaches involved integrating the need for heat in the MEA stripper with the needs of the CO_2 compression train and reducing temperature approaches in the lean/rich exchanger. This project built on that previous research by conducting an engineering and economic analysis of those innovative processing approaches to determine if significant cost savings could be achieved.

CONCEPTUAL APPROACH

This project used rigorous process modeling in conjunction with detailed engineering economics to assess the economic viability of the alternative process configurations developed at the University of Texas. The following sections describe the process configurations and the project approach which lead to the development of process simulations. From these simulations, heat and material balances were made that allowed equipment sizing and selection so that capital and operating costs could be developed and analyzed for each of the four cases discussed herein.

Process Configurations

The base case (also referred to as Case 1) for the monoethanolamine (MEA) process simulation used a feed gas with conditions typical of the exit from a flue gas desulfurization unit at a pulverized coal (PC) boiler power plant. 500 MW was selected as the gross plant capacity based on recent EIA-767 data, and a gross heat rate of 9,674 Btu/kWh was chosen based on recent EPRI data (EPRI, 2000). The coal composition and fuel heating value were based on guidelines from DOE for Illinois #6 coal (DOE, 2004). The conditions for the flue gas as well as the absorber and the stripper base case designs were derived from previous modeling research (Freguia, 2002). Table 1 displays the resultant gas conditions that feed the amine unit.

Table 1. Feed Gas Conditions

Description	Value	Unit	Value	Unit
Composition (mol%)				
CO_2	12.33			
H_2O	9.41			
N_2	73.49			
O_2	4.77			
Water saturation temperature	47	C	116.6	F
Absorber inlet temperature	55	C	131	F
Absorber inlet pressure	111.325	kPa	16.15	psia
Mole flow (after saturation)	0.0794	kmol/m2-s	0.0162	lbmol/ft2-s

Case 1 is shown in Figure 1. This case includes a basic MEA unit followed by compression of the CO_2 to 8.6 MPa (1250 psia), then cooling the CO_2 with water and pumping the dense phase CO_2 up to 13.9 MPa (2015 psia) for transport. Case 1 is based on 90% CO_2 removal.

Figure 2 illustrates Case 2. This case is similar to Case 1, but with two significant differences in the process. Heat recovery is achieved by eliminating the reflux condenser, compressing the entire stripper overheads stream up to 8.6 MPa (1250 psia) with multistage compression, and using the hot compressor discharge stream from each stage as a heat source for the amine reboiler. After passing through the amine reboiler and being cooled, condensate (water) is recovered from the CO_2 stream and recycled to the process, and the dense CO_2 is pumped up to pipeline pressure.

Figure 3 outlines Case 3. This case includes the heat recovery of Case 2 but with vapor recompression added into the stripper. Essentially, the stripper is modified to integrate the first two stages of compression into the stripper. All of the vapors from the stripper are compressed and reinjected at the next higher pressure as the vapor progresses up the column. The bottom of the stripper operates at approximately 202.6 kPa (29.4 psia), the middle section operates at 283.7 kPa (41.2 psia), and the top of the stripper operates at 405.3 kPa (58.8 psia).

Figure 4 provides a block flow diagram of Case 4. Case 4 is very similar to Case 1 (both the reflux condenser and compressor discharge are cooled with water and no heat recovery from the later stages of compression is integrated into the process), but Case 4 operates with a multipressure stripper as described in Case 3.

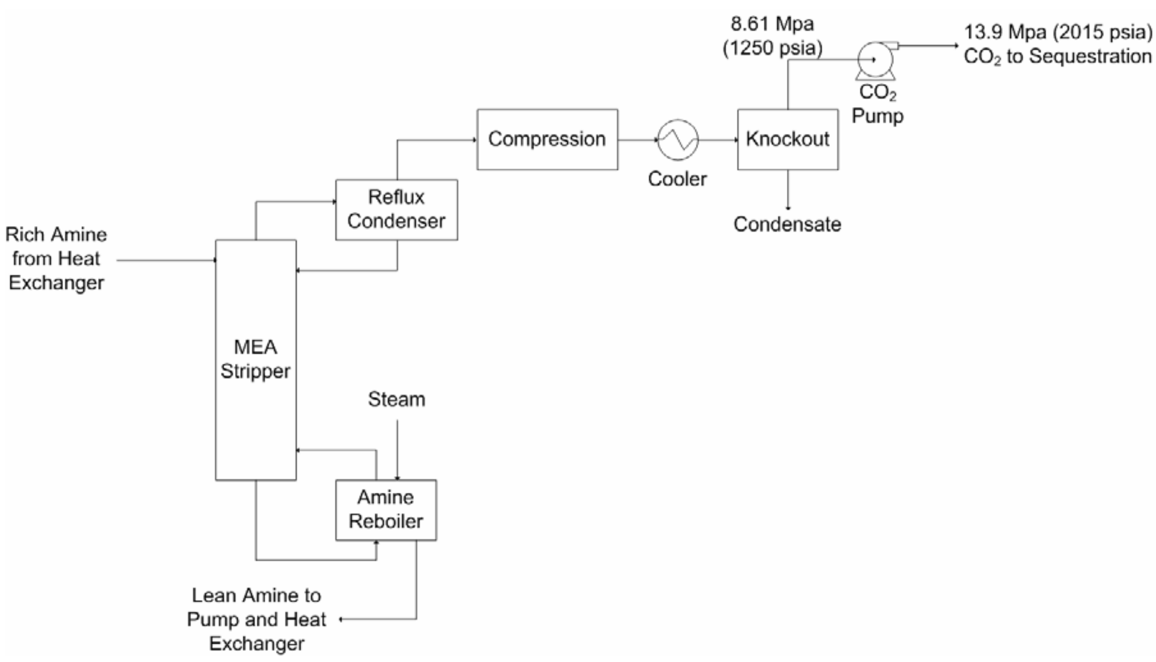

Figure 1. Case 1: Compression off MEA Stripper

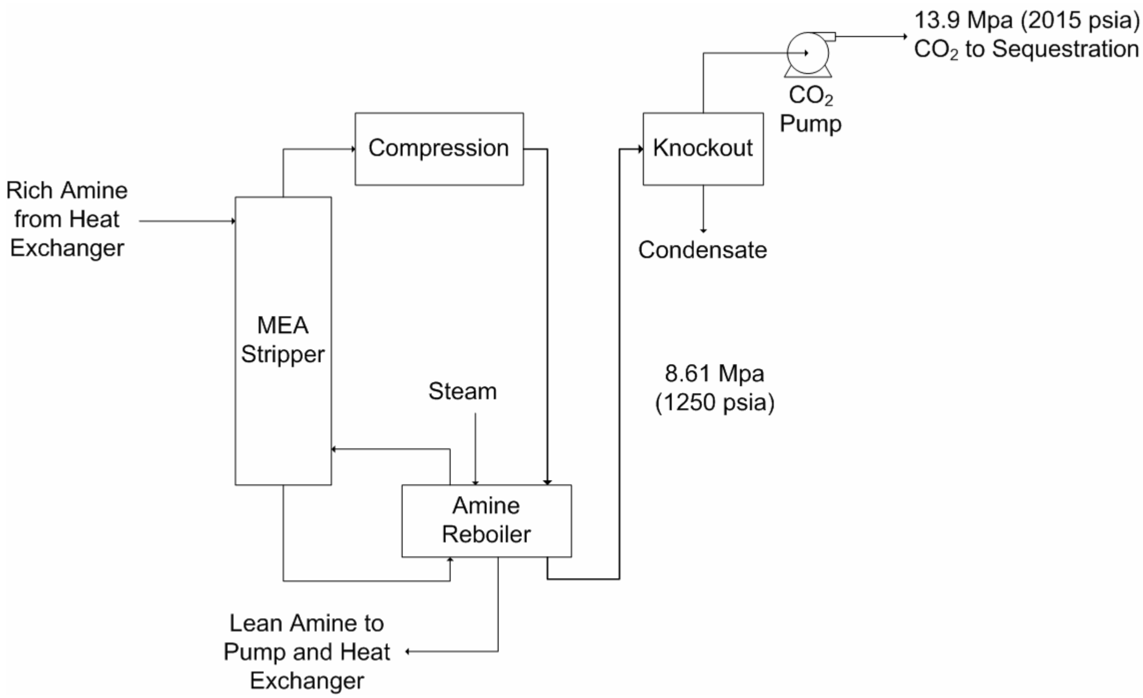

Figure 2. Case 2: CO_2 Compression with Heat Recovery

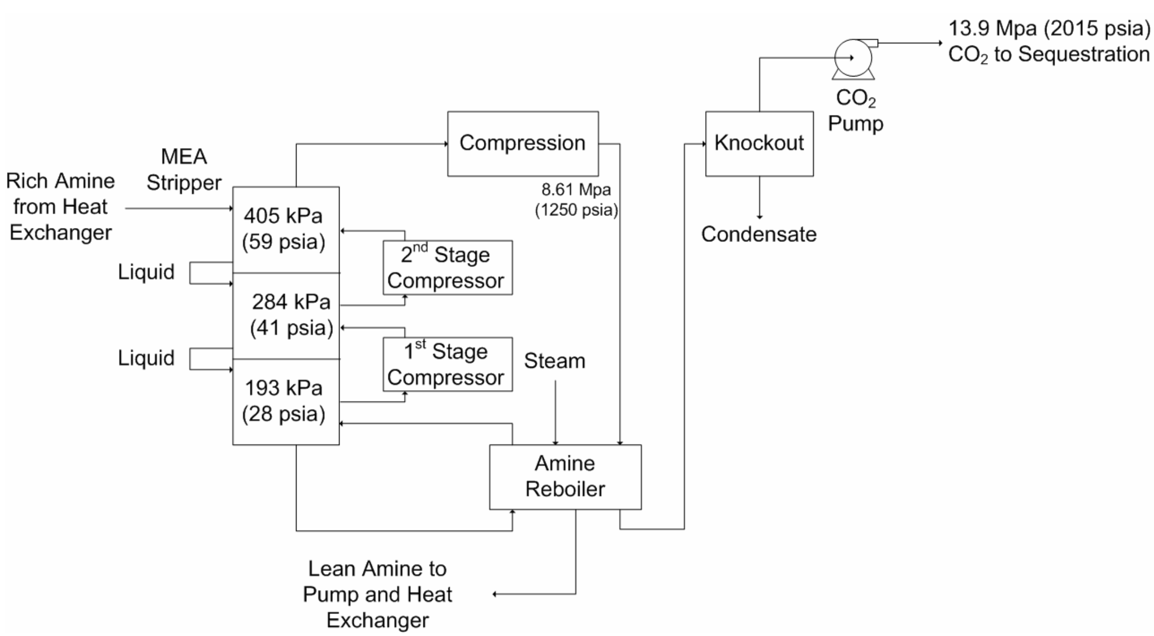

Figure 3. Case 3: Multipressure Stripping with Heat Recovery

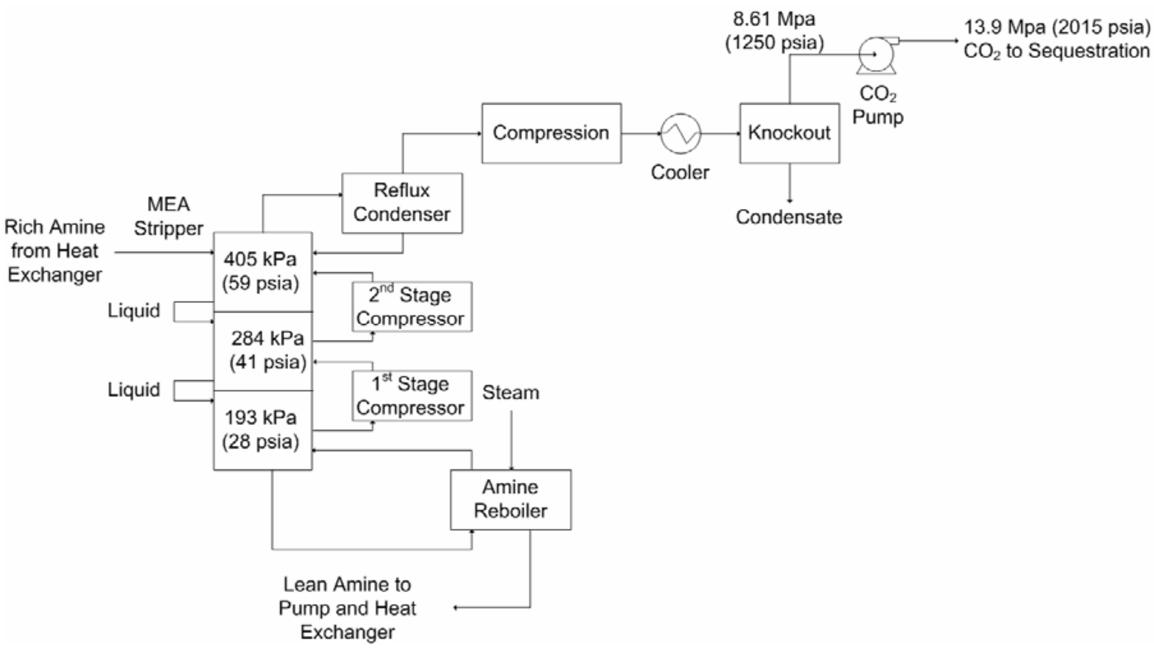

Figure 4. Case 4: Multipressure Stripper without Heat Recovery

Simulations

The University of Texas at Austin conducted process simulations using Aspen Plus with RateFrac for both the absorber and stripper. The absorber was modeled with kinetic reactions, while the stripper model used only equilibrium reactions. The Non-Random-Two-Liquid (NRTL) model for electrolyte solutions was used both for calculating equilibrium in the stripper and for calculating activities for the kinetic modeling in the absorber. The data regressed for the NRTL model were based on the work of Jou and Mather (1995). Trimeric modeled some ancillary processes, e.g., the steam desuperheating, cooling water system, and the CO_2 compression trains, using WinSim's Design II. All of the process calculations were based on steady-state conditions at the full design capacity of the unit for each case. The scope of the simulations was limited to the CO_2 capture and compression equipment. Table 2 shows the process simulation results. The primary differences between cases are the condenser duty, the net reboiler duty, the rich amine pump power, and the compressor power.

Heat and Material Balances

The outputs from both the AspenPlus and the Design II simulations were used to create heat and material balances. The material balances included detailed electrolyte species distribution.

Table 2. Summary of Process Simulation Results

		Case 1	Case 2	Case 3	Case 4
CO_2 Percent Removal	%	90	90	90	90
Amine circulation rate (lean)	gpm	41,800	41,800	42,700	42,700
Rich amine CO_2 loading	gmol/L	1.96	1.96	1.94	1.94
Lean amine CO_2 loading	gmol/L	1.18	1.18	1.18	1.18
Rich/lean heat exchanger duty	MMBtu/hr	1,750	1,750	1,750	1,750
Reflux condenser duty	MMBtu/hr	551	-	-	250
Reboiler duty	MMBtu/hr	1,710	1,710	1,330	1,330
Net Reboiler duty	MMBtu/hr	1,710	1,040	1,090	1,330
Lean cooler duty	MMBtu/hr	979	979	1,020	1,020
Rich amine pump power	hp	2,330	2,330	3,130	3,130
Lean amine pump power	hp	1,550	1,550	1,560	1,560
CO_2 compressor stages required*	-	4	9	8	5
CO_2 compressor power	hp	46,700	94,600	83,700	65,200
CO_2 pump power	hp	1,340	1,330	1,340	1,330

* *Excludes CO_2 pump stage*

Equipment Sizing and Selection

A combination of spreadsheet calculations and simulation tools (Aspen Plus, Design II, and PDQ$) were used to help size the equipment in the process. The basis of the study was four parallel amine units followed by a common downstream compression system. The following equipment items were considered:

- Absorber
- Rich amine pump
- Solution filtration
- Stripper
- Reflux condenser
- Reflux accumulator
- Reflux pump
- Reboiler
- Rich/lean exchanger
- Surge tank
- Lean amine pump
- Lean amine cooler
- Amine storage tank
- Amine makeup pump
- Water makeup tank
- Water makeup pump
- CO_2 compressors
- CO_2 interstage coolers
- CO_2 pump
- Cooling tower system
- Flue gas blower
- Reclaimer

Several assumptions were used to size and select equipment.

- The absorber and stripper were both packed towers.
- The absorber maximum diameter was set at 32 ft.
- Segments of the multipressure strippers have different diameters.
- The rich/lean exchanger uses a 10°C approach on the hot end of the exchanger.
- The lean loading for all of the cases was optimized, and the minimum total work was achieved when the lean loading was 0.25 mol CO_2/mol MEA.

- The compressors were centrifugal.
- Steam was used to drive the compressors and then heat the reboilers. In cases where the amount of steam required for the reboiler was not enough to drive the compressors, the remaining compressor load was provided with an electric motor using electricity produced by the power plant.

The majority of differences in the equipment size requirements from the base case (Case 1) occurred between the interactions of the reflux system and reboiler duty requirements, compression interstage cooling requirements, and compression work for the various flow schemes.

Capital and Operating Costs

The purchased equipment costs for the amine unit and downstream compression train were obtained from a combination of vendor quotes and costing software. PDQ$ (Preliminary Design and Quoting Service) is a software package that can be used to estimate current purchased equipment costs for chemical process equipment. The costs are in September 2004 dollars. The software estimates costs for fabricated equipment and catalog items that are based on vendor information.

The major differences in cost are related to the cost of the compressors and steam/electric drivers and the tradeoffs in where the heat exchange in the process takes place. In general, the reflux condensers required less surface area for heat transfer than the compression interstage coolers because of their higher heat transfer coefficients. The same is true with the reboiler, where straight steam requires less heat transfer area than when using the process gas from the vapor recompression interstages.

The installed costs for purchased equipment (everything but compression) was estimated using typical factors for percentage of purchased cost as reported in chemical engineering literature (Peters and Timmerhaus, 1980). The installed cost factor for compression was based on vendor recommendations for this type of application.

Engineering/home office, project contingency, and process contingency were then added to the total process plant cost to arrive at the total plant cost (TPC). Interest during construction was added to arrive at the total plant investment (TPI). The total capital requirement (TCR) is the summation of the total plant investment and prepaid royalties, startup costs, spare parts, working capital, and land.

Operating and maintenance costs included the following:

- Total maintenance cost,
- Maintenance cost allocated to labor,
- Administration and support labor cost,
- Operating labor,
- MEA cost,
- Water cost, and
- Solid waste disposal cost.

Once the total capital requirement (TCR) and the total O&M costs are known, the total annualized cost of the power plant was estimated using a capital recovery factor of 15%.

ECONOMIC ANALYSIS AND RESULTS

Table 3 presents the cost of electricity once CO_2 recovery is added for each of the four 90% recovery cases. As shown in the table, the cost of electricity is highest ($63.2/MW-hr) for Case 1, which is the conventional MEA system with compression of the CO_2 after the reflux condenser. Case 2, which removes the reflux condenser and incorporates heat recovery, has an electricity cost of $61.8/MW-hr, which represents a 2.2% savings. Case 3, incorporating both heat recovery and multipressure stripping, has the lowest cost at $59.9/MW-hr, resulting in a savings of about 5.2% over Case 1. Finally, Case 4, including multipressure stripping without heat recovery, has a cost of $60.3/MW-hr; this is approximately a 4.6% savings over Case 1. In evaluating the savings in the cost of electricity, the base coal-fired plant costs comprise a significant portion of the overall cost of electricity; this relatively fixed portion makes the cost savings appear smaller than if they were evaluated on just the cost of CO_2 capture

Table 3. Cost of Electricity with CO_2 Removal Equipment Installed

	No CO_2	Case 1	Case 2	Case 3	Case 4
Power Plant size, MW	500	500	500	500	500
Net Power Production (after power plant aux. and CO_2 capture)	453	281	294	298	293
CO_2 Recovery, tonne/hr		415	415	415	415
Power Plant Cost, $/MW-hr	25				
Annual Power Plant Cost, MM$/yr	84.3	84.3	84.3	84.3	84.3
CO_2 Removal Plant Variable O&M Costs, MM$/yr		7.6	7.0	7.0	7.3
CO_2 Removal Plant Fixed O&M Costs, MM$/yr		5.2	5.7	5.4	5.2
CO_2 Removal Plant Capital Recovery * Costs, MM$/yr		35.0	38.2	35.9	34.9
Total CO_2 Removal Costs, MM$/yr		47.9	50.9	48.4	47.5
Total Power Plant + CO_2 Removal Costs, $/yr		132	135	133	132
Cost of Electricity, $/MW-hr		63.2	61.8	59.9	60.3

** 15% capital recovery factor*

Table 4 illustrates the cost of CO_2 avoidance for the four cases. As shown in the table, the base cost of CO_2 avoidance for Case 1 is $44.9/tonne CO_2. The integration of heat recovery in Case 2 achieves a 4.6% reduction in the cost of CO_2 removal, while the addition of the multipressure stripper in Case 3 creates a cost savings of 9.8% over Case 1. Case 4, which includes the multipressure stripper without the heat recovery, leads to a cost savings of 8.4%.

Table 4. Summary of Cost of CO_2 Avoidance for a Gross 500 MW Coal-Fired Power Plant

	Units	Case 1	Case 2	Case 3	Case 4
Net power plant output after derating (base plant[1] and CO_2 capture and compression)	MW	280	293	297	293
Reduction in net power rating due to parasitic loads	%	38.1	35.2	34.4	35.3
Base cost of electricity (without CO_2 capture)	$/MW-hr	25	25	25	25
Cost of electricity with 90% CO_2 capture	$/MW-hr	63.2	61.8	59.9	60.3
Base emissions (without CO_2 capture)	MMtonnes /yr	3.43	3.43	3.43	3.43
	tonnes/ MW-hr	1.016	1.016	1.016	1.016
CO_2 Emissions with CO_2 capture (90% removal)	MMtonnes /yr	0.34	0.34	0.34	0.34
	tonnes/ MW-hr	0.164	0.157	0.155	0.157
Cost of CO_2 avoidance	$/tonne	44.9	42.8	40.5	41.1
Reduction from Case 1	%	--	4.6	9.8	8.4

Note 1: Base plant includes electricity for PC, ESP, FGD, and SCR systems for 500 MW unit (500 MW Gross, 453 MW Net)

CONCLUSIONS

The major conclusions of this work are summarized in the following paragraphs:

- Reductions in the cost of CO_2 capture ($/tonne CO_2 avoided) ranged from 4.6 to 9.8 percent among the cases;

- The configuration with the least cost per tonne avoided was Case 3 (multipressure stripping with vapor recompression);

- The parasitic energy load (as defined by the difference in net power production before and after CO_2 capture/compression equipment is installed) could be reduced by 8 – 10 percent, freeing up 13 – 17 MW of power for sale to the grid based on the model 500 MW (gross) power plant used in this study;

- The value of this incremental increase in net power production results in a short payback on capital, approximately six months to one year for Cases 3 and 4 (assuming a value of 0.06 $/kWh), suggesting that these heat integration processes are very likely to be implemented at future CO_2 capture facilities using MEA; and

- Reboiler steam requirements were reduced by 22 to 39 percent, which is desirable from the utility operating perspective despite the partially offsetting increases in electrical requirements for the compression train.

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